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DEVELOPMENT OF A BENCH-SCALE FLUIDIZED BED COMBUSTOR FOR COAL REACTIVITY STUDIES

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MISE AU POINT D'UNE UNITÉ DE COMBUSTION EN LIT FLUIDISÉ À L'ÉCHELLE DU LABORATOIRE POUR L'ÉTUDE DE LA RÉACTIVITE DES CHARBONS

par

I.T. Lau*

RÉSUMÉ

On a mis au point une unité de combustion sur lit fluidisé à l'échelle du laboratoire, en vue de coter la réactivité apparente de divers charbons. Cette réactivité apparente est une fonction complexe de nombreuses propriétés du charbon et est une mesure de la vitesse globale de combustion plutôt qu'une réaction exclusive de carbonisation en surface.

Lors de la mise au point de méthodes expérimentales et de modes opératoires destinés à mesurer la réactivité, on a effectué des mesures préliminaires avec un mélange de charbon de Pennsylvanie. Les résultats ont révélé que la vitesse de combustion dépend beaucoup de la taille des particules de charbon, ce qui est en accord avec le modèle de la sphère à diamètre décroissant. On a aussi étudié l'effet d'autres paramètres fonctionnels, comme la température du lit, la teneur en oxygène et la vitesse du gaz de fluidisation.

Afin d'améliorer le rendement, des modifications seront apportées au système d'enregistrement et au système d'alimentation en charbon; il sera aussi nécessaire de mettre au point une nouvelle expression pour la vitesse en vue de déterminer la réactivité apparente durant la combustion en lit fluidisé.

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DEVELOPMENT OF A BENCH-SCALE FLUIDIZED BED COMBUSTOR FOR COAL REACTIVITY STUDIES

by

I.T. Lau*

ABSTRACT

A bench-scale FBC unit has been developed to rank the apparent reactivity of various coals when burned by fluidized bed combustion. This apparent reactivity is a complex function of many of the coal properties and is a measure of the overall combustion rate rather than the char surface reaction only.

In developing experimental methods and procedures for the reactivity measurement, preliminary runs were conducted with a blended Pennsylvania coal. Results show that the combustion rate is highly affected by the coal particle size which is consistent with the shrinking sphere model. The effect of other operating parameters such as bed temperature, oxygen level, and fluidizing gas velocity was also investigated.

The recording system and the coal feeding system will be modified to improve performance and it will also be necessary to develop a new rate expression as a measure of the apparent FBC reactivity.

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INTRODUCTION

The technology of atmospheric fluidized bed combustion (AFBC) of coal is now well established. Its advantages have been demonstrated in many bench and pilot units in recent years and industrial units are now being offered in the market (1,2). One of the most important advantages is the ability of FBC systems to burn a much wider range of coals and solid fuels than any given conventional burning technology.

In Canada, coal mines produce various ranks of coals from lignite to anthracite, including many unreactive oxidized coals. A major objective of the FBC program is to provide sufficient background information on the FBC characteristics of various coals. The performances of Canadian coals in a FBC environment are currently being studied at CANMET by means of a pilot-scale unit (3,4,5,6). To reduce the amount of pilot-scale work necessary to characterize the combustion behaviour of Canadian solid fuels, bench-scale FBC work to obtain kinetic and reactivity data on these fuels is desirable. Such information would also be useful in planning pilot-scale research. With this in mind a bench-scale fluidized bed combustor has been designed and built for coal reactivity studies at the CANMET Combustion and Carbonization Research Laboratory. This small unit has also been designed to provide the simplicity and flexibility to perform fundamental FBC studies and to test new ideas as they arise.

INTRINSIC REACTIVITY, SURFACE REACTIVITY, AND APPARENT FBC REACTIVITY

Combustion of coal involves burning of volatiles and char. The reaction mechanisms of the combustion of these two materials are different. In general the char reaction time is about two orders of magnitude greater than that of volatiles. For simplicity most reactivity studies focus on char reactivity and neglect volatiles.

The combustion of carbon by oxygen is best described by the shrinking core model. However, because char is a highly porous material, a slight degree of internal burning in the solid has been observed (7). This means that the reaction actually takes place in a thin layer at the char surface.

The degree of oxygen penetration before reaction with the carbon depends on temperature, intrinsic reactivity, particle size and pore size of the char. Intrinsic reactivity is defined as the reaction rate per unit pore surface area in the absence of any restriction due to diffusion (8). In practice, the intra-particle processes of pore diffusion and chemical reaction are usually lumped into a single rate expression which is defined here as apparent surface reactivity based on the external surface area of the particle.

Coal char reactivities have been investigated using several different methods (8,9). Recently, fluidized bed reactors have been suggested as a tool for routine reactivity measurement (9) to take advantage of the high heat transfer and uniform materials temperature characteristics of such systems. However, the effects of the fluidized bed's hydrodynamics and heat transfer introduce additional uncertainties to the complex problem of elucidating the mechanisms of coal combustion.

Many interrelated physical and chemical processes such as fragmentation and attrition of solid fuels, elutriation of unburned carbons, mass transport and heat transfer, heterogeneous and homogeneous reactions, in a fluidized bed reactor may influence the overall combustion rate. Apart from the intra-particle processes, other process factors must be eliminated if the surface reactivity is to be obtained from the system's rate data. The effects of pore diffusion and surface reaction must also be distinguished for intrinsic reactivity measurement. However, if the kinetic information obtained is going to be applied directly to FBC systems, it is more convenient to lump all these process effects into a single rate expression. Such an expression, which is a measure of the overall coal combustion rate rather than of the char surface kinetics in a FBC, is defined here as the apparent FBC reactivity. Since the original nature of the coal is an important factor in influencing various processes including devolatilization, fragmentation, and attrition, coal rankings on the basis of FBC reactivity and surface reactivity may be different.

The main objectives in building the bench-scale FBC described in this work are to measure and to rank various Canadian coals in terms of apparent FBC reactivity. Of course, by varying the operating conditions, the apparatus is also expected to obtain other forms of reactivities.

REACTION MECHANISMS

To develop a process to evaluate coal reactivities in a FBC, the reaction mechanisms of the system must be well understood. There are many models available in the literature describing these complicated phenomena. Some of the more important work has been reviewed recently by Park et al. (10) and subsequently by LaNauze and Jung (7). Although there are conflicting views, the following most generally accepted concepts are assumed in this study.

COMBUSTION OF VOLATILES

When coals are fed into a hot fluidized bed, the particle temperature rises rapidly by absorbing heat from the surrounding inerts. Volatile matter evolved from the coal ignites and burns instantaneously under diffusion controlled conditions.

The subsequent char reactivity and porosity depend on the devolatilization process. Some coals having caking properties may swell during volatiles evolution. Fragmentation and attrition are expected to be strongly influenced by the structure and toughness of the char, hence by the devolatilization process.

The volatiles release time increases with particle size (11,12) and is comparable to the solid circulation time. In considering most of the coal size used in FBCs, the volatile flux is large enough to prevent oxygen from reaching the solid surface. The combustion steps of volatiles burning and char combustion can be treated independantly.

GAS AND SOLID MIXING

The performance of a fluidized bed combustor is strongly influenced by the fluid mechanics of the bed. Most of the early theories developed employed Davidson and Harrison's two phase theory (13). This assumes that the excess gas flow beyond that required for minimum fluidization passes through the bed in bubbles. Bubbles grow due to coalescence and bring solids in the wake to the top of the bed. Solid particles are considered to be well mixed whereas the gas in bubbles is regarded as being in plug flow.

In recent years the differences between large and small size particle fluidization have been noted (14). The FBC operates practically in the so-called slow bubble regime where the bubble rise velocity is slow relative to the interstitial gas velocity. Since no gas circulates back into them, the bubbles are cloudless and the interstitial gas uses them as a by-pass. The interstitial gas can also be assumed to pass through the bed in plug flow when the magnitude of its velocity exceeds the velocity of the downward flowing solids (15).

CARBON-OXYGEN REACTION

The heterogeneous reaction between oxygen and carbon is assumed to occur in the dense phase and the wake of bubbles only. Within these regions, oxygen transfers through gas resistance film to the char surface where the following reactions may take place:

$$2C + O_2 \rightarrow 2CO \qquad (1)$$

$$C + O_2 \rightarrow CO_2 \qquad (2)$$

$$C + CO_2 \rightarrow 2CO \qquad (3)$$

Equation 3 of the C-CO_2 reaction was found to be too slow at FBC temperatures to give significant contribution to the overall reaction (16). Recent studies show that carbon monoxide is the primary product at the char surface and that the CO diffuses away from the particle and is then converted to CO_2 (17,18).

The distance at which the CO-CO_2 conversion occurs depends on fluid mechanics, heat transfer and particle size. The location of CO flame (whether within or after the resistance film) has a major impact on the char burning rate when the rate is external mass transfer controlled.

On the burning char surface, residue ash is flaked off and presents no resistance to the reaction. Another contribution to the oxygen diffusion resistance, which may influence the overall reaction rate, is given by the bubble phase to dense phase mass transport.

The contribution of the various rate controlling steps is a complicated function of bed temperature, fluid mechanics, oxygen concentration, particle size and char surface reactivity. In general, external mass transfer and surface reaction are both important for burning coal in a FBC. As the

mass transfer coefficient is inversely proportional to the particle size, the combustion of large char particles is dominated by mass transfer control whereas chemical kinetics become important for the burning small diameter particles.

ATTRITION, FRAGMENTATION AND ELUTRIATION

One of the main drawbacks of the FBC system is the elutriation of combustible fines out of the reactor which in turn causes low combustion efficiencies. It has been reported that the main source of elutriated carbon is the fines generated by the attrition of char surface by abrasion (19,20). Elutriation rates of attrited carbon are approximately proportional to the product of the excess of gas velocity above the minimum for fluidization, i.e., $(U-U_{mf})$ and the overall carbon surface exposed to attrition inside the bed (21).

Attrition is enhanced by combustion and its rate is much higher during oxidation than pyrolysis. Coals of higher reactivity burn faster and result in less carbon inventory in the bed. Consequently there is less carbon surface exposed to attrition which produces less elutriated combustible fines hence better combustion efficiency.

Thermal fragmentation does not significantly contribute to the emission of carbon fines through an increased number of unburnt particles (20). However, fragmentation and swelling of particles during devolatilization create more surface area for reaction. Thus, reaction kinetics are obviously complicated by attrition and fragmentation processes.

EQUIPMENT DESCRIPTION

CONCEPTUAL DESIGN AND BASIC PARAMETERS

To emphasize the effects of coal properties on the rate data, FBC experiments must be performed under conditions where external mass transfer effects are minimized. It is well known that employing small particle size and low temperature can move the reaction mechanisms from mass transfer towards chemical reaction control. However, many important combustion-rate-dependent effects such as devolatilization, thermal fragmentation, swelling, and even attrition may be largely suppressed under these conditions. Therefore, the major design parameters in the present work were selected to be as close as possible to those of practical FBC operating conditions.

Table 1 gives the selected basic parameters used to design the bench-scale fluidized bed rig and its accessories, as well as to estimate the requirements of gas flow, pressure drop, heating power, etc.

Because the devolatilization process is so important to the subsequent char properties, it was decided to use coal rather than char in this work. In practice, coal feeds used in FBC have a wide size range. The shrinking sphere model predicts that the reaction rate is in proportion to particle diameter raised by a power of 1 to 2. It has been noted that coals of various sizes may have different reactivities after crushing. Also, from the devolatilization point of view, the properties of char produced by a large piece of coal may not be identical to those of smaller pieces. Thus the determinations and the comparisons of apparent FBC reactivities from different coals should be based on the size ranges.

In this work, a batchwise operating system was chosen to take advantage of the simplicity of operation and the possibility of obtaining a direct record of the carbon burning history. To reduce fluctuations of bed temperature and oxygen concentration, the size of coal sample was restricted to maintain coal-inert ratio in the bed under 2%.

Silica sand (mean size 0.5 mm) was chosen as the bed material. Notwithstanding its smaller particle size when compared with that used in most practical FBCs (~1 to 2 mm) it has the advantage of higher heat transfer rates - an important factor for reactivity studies.

It is difficult to reproduce gas and solid flow patterns experienced in other unit beds especially when their sizes are significantly different. A common problem in small-scale fluid bed units is that one tends to operate in slugging rather than bubbling regime, thus the fluidizing characteristics change. The aspect ratio (bed height to bed diameter) restriction must be precisely controlled if operation is to avoid slugging. From Yagi and Muchi's correlation, the maximum aspect ratio allowed was estimated at 1.8 for the basic parameters required for this study (22).

The amount of sand to be used is determined by the choice of the rig size. A smaller size unit has the advantage of requiring less fuel, air, heat and power. However, slugging and wall effect are likely to be serious problems when fluidized bed column's diameter is too small. Consequently, a value of 0.1 m was chosen as bed diameter which gives the maximum fluidized bed height a value of 1.8 m from the previous estimated aspect ratio.

The equation proposed by Darton et al. from the theory for bubble growth due to coalescence can also be used to estimate the maximum bed height (23).

$$h = \frac{d_b\, g^{1/4}}{0.54\,(U-U_{mf})^{1/2}} - 4\,A_o^{1/2}$$

where h is the height of the bubble above the distributor; d_b is the bubble diameter; U is the superficial gas velocity; U_{mf} is the minimum fluidization velocity; and A_o is the area of distributor plate per orifice.

For the present work, a value of 1.5 m/s was chosen as the maximum fluidizing gas velocity (Table 1). This value is similar to that used in general FBC practice although it is considered high for the particle size used in this study.

The value of U_{mf} can be calculated at about 0.1 m/s for the present conditions and A_o, determined from the gas distributor design (described later), is equal to 1.51×10^{-4} m^2. Thus the maximum value of fluidized bed height calculated from Darton et al. is 0.23 m which is in fairly good agreement with that from Yagi and Muchi (22).

The value of 0.18 m was therefore taken as the maximum possible height if slugging was to be avoided. The appropriate amount of sand required for the bed was estimated to be 1 kg assuming a 0.75 fluidized bed void fraction. Depending on the conditions required, coal samples of 20 g or less may be used to keep the coal-inert ratio low enough, i.e., less than 2%.

When feeding small coal samples and using air as the fluidizing gas, the oxygen level in the present batch system operation will approach 21% which is much higher than that experienced by most FBC systems. The main reason for not using nitrogen and air mixes to maintain more usual excess oxygen levels is the requirement to observe a volatile flame, allowing the processes of devolatilization and char burning to be decoupled. Another reason for keeping a high oxygen content in the bed is that the mass transfer effect from bubble phase to dense phase can be minimized.

The overall column height and the location of cyclone can be determined by the evaluation of the value of transport disengaging height (TDH) which is the height above the bed where the superficial velocity profile has become stable. Above the TDH the gas can only carry its saturation capacity

of those particles of the bed having a terminal velocity less than the superficial gas velocity. The TDH value obtained from Zenz and Weil is about 0.7 m for the highest fluidizing gas velocity used in this study (24).

THE FLUIDIZED BED COMBUSTOR

Three sections comprise the bench-scale unit: windbox, fluidized bed, and expanded freeboard. These sections are bolted together and can be extended if required. The overall dimensions are given in Table 1. The rig is fabricated of number 310 stainless steel to ensure adequate performance approaching 1000°C.

In the fluidized bed section, ports at various heights are provided for temperature and pressure measurements. Pressure probes are inserted at an angle of 60° to the horizontal to prevent solids blockage. All thermocouples and pressure probes can be adjusted freely to the desired positions.

A viewport-feeder system which is fabricated of 63.5-mm diam steel pipes is shown in Fig. 1. The viewport portion is located just above the maximum fluidized bed height and is welded on the fluidized bed section at an angle of 60° to the horizontal.

A vertical feeder is affixed to the viewport portion which consists of two ball valves. A pneumatic line supplies an air jet to clean coal particles into the bed.

A solenoid valve is used to direct air stream either to the feeder or to sweep the dust off the viewport glass. The solenoid valve is controlled by an electric timer. The flowrate of the pneumatic line is controlled by a needle valve and is monitored by a rotameter.

Although the unit is to be used mainly for batchwise operations, a 25-mm diam continuous feeding line is provided in the fluidized bed section to allow continuous feeding if desired.

The rig diameter is enlarged to 0.2 m at the expanded freeboard section. The top of this section has a hinged hatch to allow the removal of bed material. Two ports located on the hatch are provided for freeboard cooling coils or sampling ports. The flue gas exit port is located at the side of this section and is directly connected to a cyclone.

Fluidizing gas is supplied through a side port at the windbox section. A solid withdrawal pipe is located at the bottom of this section to control bed inventory if a continuous operating system is used.

Between the windbox and the fluidized bed section lies a perforated stainless steel plate which distributes gas evenly into the bed. Two criteria were followed in the gas distributor design. First, the flowrate through the distributor plate was determined according to the minimum pressure drop requirement suggested by Kunii and Levenspiel (15). Secondly, the distributor plate was designed to follow Zenz's recommendation that the grid hole spacing should not be less than 1.5 times the bubble diam at the level of the penetration jet height (25). The calculated result suggested the distributor plate should be fabricated with 52 0.8-mm diam holes arranged in a triangular pitch. A 44 μm mesh size stainless steel screen was welded to the bottom of the distributor plate to prevent weeping of bed solids through the orifices. The design of the distributor assembly thus allows easy replacement of the plate if necessary.

FLUIDIZING GAS INLET AND FLUE GAS EXHAUST SYSTEMS

A schematic of the gas flow system is given in Fig. 2. Either air, nitrogen or a mixture of these can be supplied as the fluidizing gas into the combustor. An electrically operated three-way solenoid valve allows the sudden exchange of the inlet gas. Besides permitting regular combustion experiments, such an arrangement allows one to perform studies of in situ devolatilization followed by char burning and, intermittent combustion by allowing rapid cooling of the bed.

To reduce the heat load at the fluidized bed section, inlet gas is preheated to about 700°C. The required length of a 13-mm diam stainless steel tubing for preheating gas from room temperature to 700°C was calculated from the forced convection heat transfer correlation.

With 30% allowance added to the resulting value, a 1.37 m long steel tubing at the end section of the gas line was wrapped around the windbox section which is enclosed by electrical heaters.

The flue gas is emitted to the atmosphere via a cyclone. The dust collecting efficiency of a cyclone depends on its gas inlet velocity. High velocity provides a large centrifugal effect to settle solid particles. However, at gas rates larger than 2 times the saltation rates, fines can be picked up and re-entrained into the gas stream (26).

Cyclone performance correlations available in the literature are primarily based on large cyclones having diameters over 0.3 m. The size of

the cyclone was calculated at 63 mm based on the gas inlet velocity being corrected to achieve an equivalent centrifugal force to that of a large unit. The other dimensions of the cyclone were determined according to criteria given in the literature (26,27).

A long steel pipe is located downstream of the cyclone so that flue gas can be cooled below 200°C before entering a fan exhaust system. To ensure positive back pressure in the fluidized bed column, a ball valve was assembled at the exhaust pipe to restrict gas flow. This is important to avoid significant errors in determining the composition of the flue gas when sampling in the event the system develops a small leak.

HEATING AND TEMPERATURE CONTROLLING

The use of an electrically heated system for a small-scale FBC unit is particularly convenient in that temperatures of the various sections can be controlled easily.

The present system involves four heating elements which are individually controlled. They require different power according to their functions. The first two heaters for the upper and lower freeboard sections need less than 0.5 kW each to maintain the desired temperature which is usually less than 750°C to suppress freeboard burning.

The results of energy balance calculations show that the third heater, located at the fluidized bed section, requires more than 3 kW to heat and maintain the temperature of the bed material whereas the windbox heater for the fourth section requires at least 2.5 kW to preheat the incoming gas. Because of the limited column surface area available for the last two sections, their Watt densities are high.

The original design used Kanthal A-1 resistant wires for all four sections. Kaynex high temperature cement was used to hold the heating elements in place and to insulate them from the metal surface. Unfortunately the thermal stress was too high for both the heating wire and the cement. The heaters were ground out and created a hot spot on the rig during the early stage of the preliminary tests.

Both the rig and the heating system were modified to allow Thermcraft RH type ceramic heating blocks to be used for the two high Watt density sections. These heating blocks were fabricated to form a special shape permitting the extrusions of the feeder-view port assembly and measuring ports.

Kanthal A-1 wires with ceramic beads to protect them from grounding were used for the upper and lower freeboard sections. Fiberfrax from Carborandom Co. was wrapped around all sections (about 8 mm thick) to provide thermal insulation.

Thermo Electric model 100 on-off and proportional outputs temperature controllers are used to control heaters on the freeboard sections. The heating elements of the bed and the windbox sections are controlled by Thermo Electric model 940 and model 910 adjustable PID controllers respectively in conjunction with a Halmar model LZF1-2425 single phase zero firing power controller.

GAS SAMPLING AND RECORDING SYSTEMS

The flue gas samples to be analyzed are continuously drawn during experiments from the downstream of the cyclone by an electric vacuum pump. A schematic of the sampling gas flow system is given in Fig. 3.

The gas sample is first dried by removing the moisture in a double condenser. The condenser is immersed in a cooling water bath where temperature is maintained at 10°C by a U-Cool immersion cooler. Heating and insulation are provided to the line upstream of the condenser to prevent line blockage due to condensation. Particulates which escape the cyclone are then removed by an in-line wool filter located downstream of the condenser.

The remaining moisture in the gas is chemically removed in a Drierite cylinder filled with anhydrous calcium sulphate. Then the gas passes through another in-line filter, to ensure it is dust free prior to entering various gas analyzers.

Two Beckman model 864 infrared analyzers are used for CO and CO_2 measurements and a Beckman model 755 paramagnetic analyzer is used for O_2 measurement. A Horiba model FIA-22A is also available for measuring the concentration of total hydrocarbons, if required.

A three-way solenoid valve in the sampling gas line enables the switch between sampling and calibration gases into the sampling system. Since the gas analyzers are pressure and flow sensitive, it is important to ensure the calibration gas flowrate is identical to the sampling gas. Fine adjustment to these rates is provided by an accurate needle valve.

To reduce the time lag, a by-pass line is branched from the sampling line to complete the whole system. The rates of gases flowing into indivi-

dual analyzers were accurately regulated by preset valves assembled upstream of analyzers and by an inline valve at the bypass and are monitored with rotameters.

The signals generated from the gas analyzers, thermocouples, and pressure transducers, etc., are recorded with a Molytek model 2700-32 channel microprocessor and two Hewlett-Packard model 7100 strip chart recorders.

PRELIMINARY EXPERIMENTS

To evaluate the equipment's performance and to develop experimental procedures for evaluating the apparent FBC reactivities of various coals, a preliminary series of combustion runs was conducted with coal amount, coal size, fluidizing gas velocity, and bed temperature as variable parameters.

MATERIALS AND PROCEDURE

A blended Pennsylvania coal was used in these runs. The proximate and ultimate analyses are given in Table 2. After crushing, the coal was screened to five size ranges which are also given in Table 2.

The fluidized bed of inert sand particles was preheated electrically with the air flowing through it. The fluidizing gas flowrate was adjusted periodically to compensate for the effects of the gradual changes of the volume expansion and viscosity of the air as the bed temperature rises. Temperature and pressures at various positions were monitored to ensure the bed was well fluidized and thoroughly mixed. Gas analyzers were calibrated with gases of known compositions.

A weighed coal sample was fed into the preheated bed zone through the pneumatic double valve feeder, the flue gas compositions and bed temperature being continuously recorded until the CO and CO_2 concentrations fell back to zero. During the early stage of combustion, the burning of volatiles can be observed through the view port. Other operating parameters such as freeboard temperature, pressure drops across the bed and cyclone were also recorded by the microprocessor.

RESULTS AND DISCUSSIONS

Typical flue gas compositions of burning coals of large particle size (over 3 mm diam) are shown in Fig. 4a. It has been noted that as char

burning commences, the CO concentration increases gradually up to a certain value while the CO_2 concentration drops continuously. The CO level diminishes afterwards at a much slower rate and, as the carbon in the bed is burned out, remains at a small asymptotically decreasing level for a longer time than the CO_2. The reason for this is not completely clear. Theoretically CO will react with O_2 to from CO_2 near the char surface for large particles but will burn outside the boundary layer for small particles. Probably as the particle size diminishes, the amount of CO formed at the char surface is reduced but so is the proportion of CO burnt to CO_2 because oxygen diffused through the film will react preferentially with carbon to form CO rather than reacting with CO. The CO concentration thus goes through a maxima and decreases more slowly than that of the CO_2.

The above argument may also be part of the reason for the higher CO proportion that results when burning coals of small particle size (under 3 mm diam) as shown in Fig. 4b. Another explanation for the greater CO concentration in the flue gas is a lower oxygen level in the bed due to the combustion rate which is much faster when burning smaller size coals.

Oxygen concentration varies inversely with CO and CO_2 concentrations and drops to a lower value when burning small size coal rather than large size coal as expected. The degrees of fluctuation of both the oxygen level and the bed temperature during combustion depend on the initial burning rate, particularly from the evolved volatiles. With the most severe conditions used, the oxygen level in the flue gas is lowered to 11% and the bed temperature increased by about 35°C.

Adding up the equivalent quantities of carbon from the CO and CO_2 concentrations from Fig. 4a yields another curve representing the carbon burning rate against time as shown in Fig. 5a. The curve illustrates the combustion rate of the large particle size coal as well as an obvious turning point at time, t_v. This point was found to closely match the observed "flame time" during experiments. Thus the time, t_v, is considered to be the volatile evolution time, or, the time after which char combustion commences.

Figure 5b gives the burning rate of coal particles in the small (<3 mm) size range. The curve shows that the rate of carbon reaction decreases gradually with no sudden turning point as given by the previous graph for large particles. A small size coal particle evolves only limited amount of volatile matter which is not enough to stop the oxygen from penetrating to

the solid surface. Hence, fixed carbon ignites before completion of devolatilization so that the two burning processes overlap and cannot be treated independently.

In the present preliminary studies, the carbon burn-out time, t_B, as given in Fig. 5, is used as a measure for the overall coal combustion rate to indicate qualitatively the effects produced by the various operating parameters mentioned previously.

Figure 6 shows that increasing coal sample size increases the carbon burn-out time. Larger quantities of coal in the bed consume more oxygen and hence reduces the oxygen concentration available for reaction. These curves indirectly indicate the oxygen level effect on the combustion rate.

When comparing curve No. 1 with curves 2 and 3 in Fig. 6, it can be seen that burning small particle size coals results in a higher fractional increase of burn-out time when increasing coal amount than when burning coals of large size ranges. Smaller size particles obviously react faster leaving less oxygen concentration in the bed, thus the combustion rate is more sensitive to the amount of coal in the bed for smaller particle size as expected.

If a small amount of a large particle size coal sample is fed into the fluidized bed combustor, the reaction would be too slow to cause any significant depletion in the oxygen level. The lack of sample size effect in this case is shown by curve No. 5.

The temperature effect shown in Fig. 7 presents contradictory results. As expected, when burning large particle size coals, t_B decreases with increasing temperature (curves 1 and 3), but for other operating conditions (curves 2 and 4) there is no temperature effect at all.

Burning small size coals shifts the reaction controlling mechanism towards surface reaction which is a highly temperature dependent process. However, curves 5 and 6 give a flat t_B profile with bed temperature.

The unexpected results are attributed partly to the deficiency of the pneumatic double valve feeder when handling small size particles. The common ball valves used have extruded rims at the openings, thus fine size coals do not always enter the bed even with an air jet. Modification of the feeder is necessary to feed fine size materials.

Careful examination of the carbon burning rate curves shows that the unexpected large burn-out time obtained from some experiments results from long tailings of the flue gas signal at the final stage of combustion.

The results are strongly influenced by the largest or the most unreactive particle in the coal sample. Inconsistent results are especially easily generated by experiments employing large particle size coal, because the coal sample of such runs contains only few particles. Therefore the burn-out time is less effective in representing the average rate of carbon reactio for FBC reactivity studies. A more reliable rate expression that could reflect the effects of the original properties of the coal on combustion rate must be developed.

The experimental results shown in Fig. 8 indicate a strong particle size effect on reaction rate which is consistent with the shrinking sphere model.

Another strong factor affecting the rate of combustion in a FBC is the superficial gas velocity which determines the fluid mechanics in the bed. When comparing curve 1 and 2 in Fig. 8 it can be seen that experiments with higher air flowrates require less time to burn out the coal. However, the effects of fluidizing gas velocity and oxygen level cannot be separated under the present operating conditions and that quantitative analysis of the gas velocity effect on combustion rate is impossible.

It is interesting to note that curve 4 was generated with a lower bed temperature but higher oxygen concentration conditions coincide with low gas flowrate (curve 1) at small particle size ranges and with high gas flowrate (curve 2) at large particle size ranges, in Fig. 8. The result, which is in contrast with the one given in Fig. 7, suggests the temperature effect is even more important when burning small size coals.

CONCLUSIONS AND RECOMMENDATIONS

1. A bench-scale FBC unit has been designed and fabricated. It has been tested under desired conditions with satisfactory results. However, the coal feeding system requires modification. Also, the data analysis employing chart recorders is tedious. A data acquisition system should be employed to replace the recorders. With a more convenient and automatic system, experimental work would be faster, easier and more accurate.
2. Burn-out time is not an ideal measure of apparent FBC coal reactivity. A new rate expression should be developed to rank various coals in the future.

3. Particle size has a strong effect on the reaction rate of coal combustion in a FBC which is consistent with the shrinking sphere model. Also, differences between the large and small coal particles in char and volatile combustion mechanisms have been demonstrated by experiments. This validates the conclusion that the reactivity properties of various coal sizes are not identical. Therefore, it is recommended that the measurement of apparent FBC coal reactivities should be based on various particle size fractions.
4. Combustion experiments using small size coal samples have the advantages of flat temperature profile and smaller fluctuation in oxygen level. However, small quantities of particles do not represent the coal of the particular size fraction well on a statistical basis as particle size exerts a major effect on the combustion rate. The sample size should be varied with more coal used for burning larger size particles.
5. The turning point at the carbon burning rate curve for large particle size (>3 mm diam) combustion of coal can be used to distinguish the processes of volatile burning and char combustion.
6. Preliminary experimental results show that the overall rate of coal combustion in a FBC increases with increasing oxygen level, superficial gas velocity, bed temperature, and with decreasing particle size.

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Table 1 - Basic design parameters and rig dimensions

Basic parameters		Column dimensions	
Coal		Bed diam:	0.1 m
Particle size:	0.1 - 25.4 mm	Expanded freeboard zone diam:	0.2 m
Sample amount:	max 20 g		
Sand		Overall length:	1.17 m
Particle size:	avg. 0.5 mm		
Amount:	1 kg	Cyclone size:	63 mm
Air			
Superficial gas velocity:	0.3 - 1.5 m/s		
Bed temperature:	600 - 1000°C		

Table 2 - Analysis of Pennsylvania coal sample

Proximate analysis	Ash	Volatile matter	Fixed carbon (by diff.)
wt% (Dry Basis)	19.24	23.32	57.44

Ultimate analysis	Carbon	Hydrogen	Sulphur	Nitrogen	Ash	Oxygen (by diff.)
wt% (Dry Basis)	65.25	1.79	0.36	1.25	19.24	12.11

Particle size fractions	1	2	3	4	5
(mm)	0.595-0.841	0.841-3.360	3.360-5.66	5.66-12.7	12.7-25.4

Fig. 1 - Bench-scale fluidized bed combustor

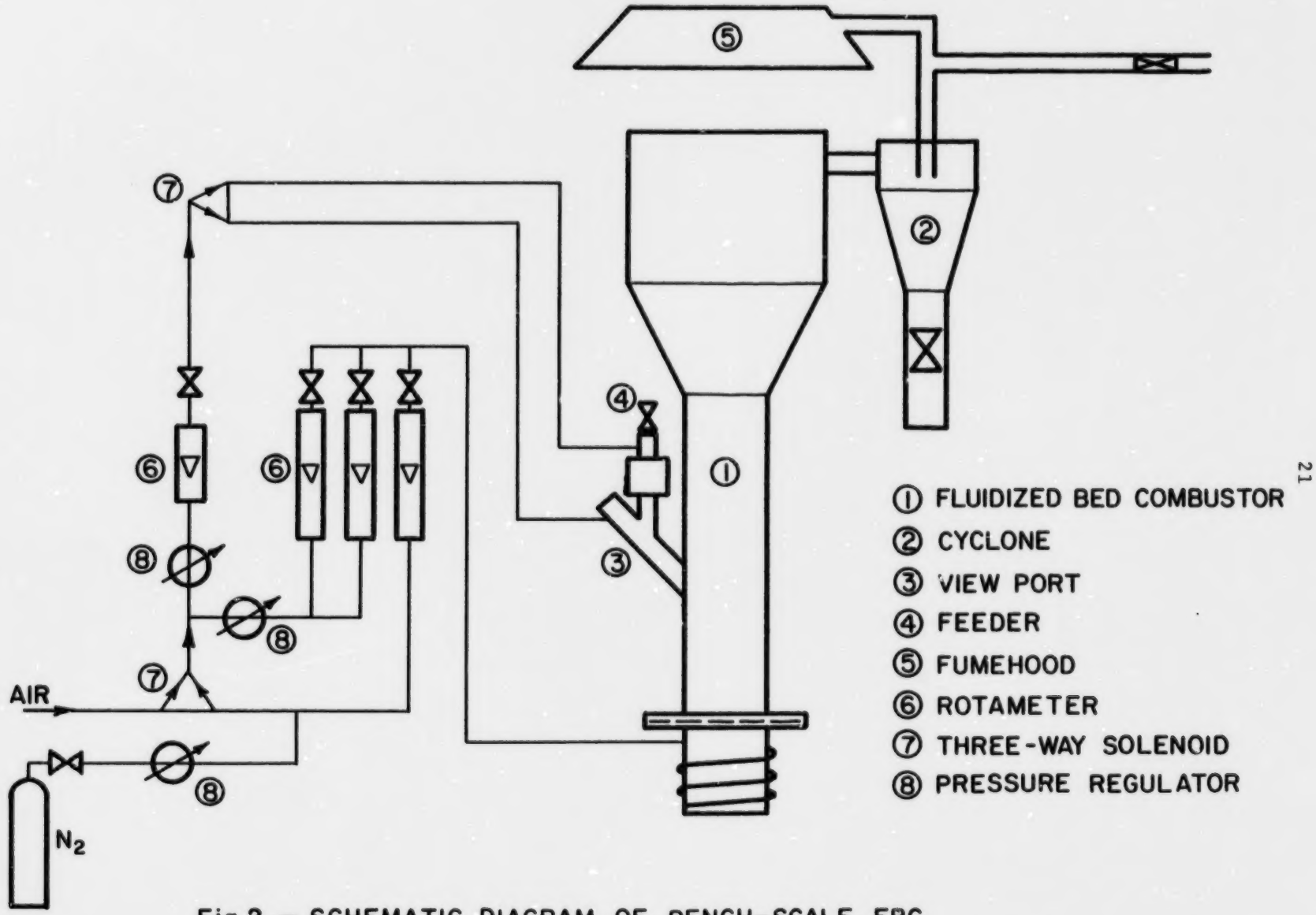

Fig. 2 – SCHEMATIC DIAGRAM OF BENCH-SCALE FBC

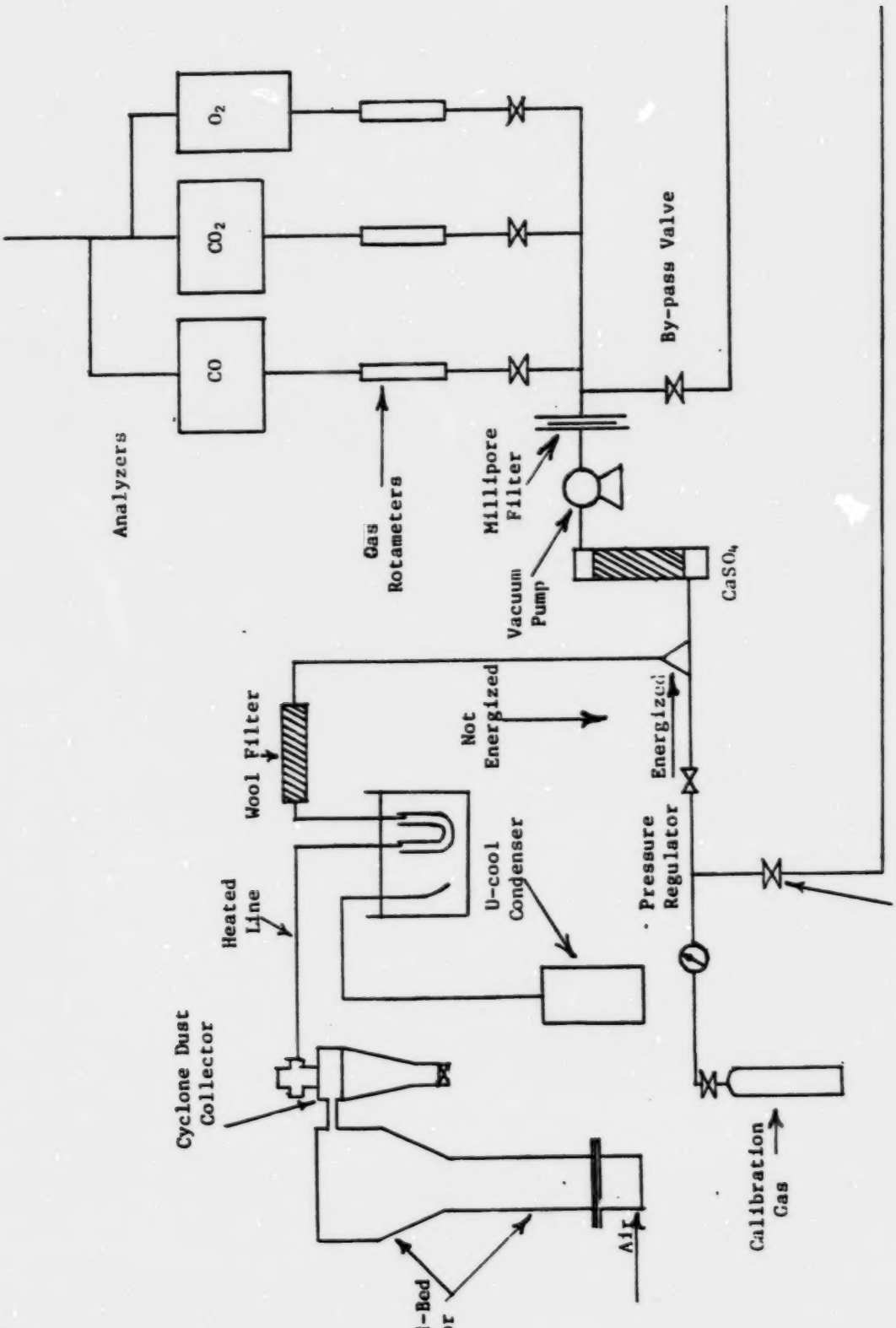

Fig. 3 - Schematic diagram of sampling gas flow

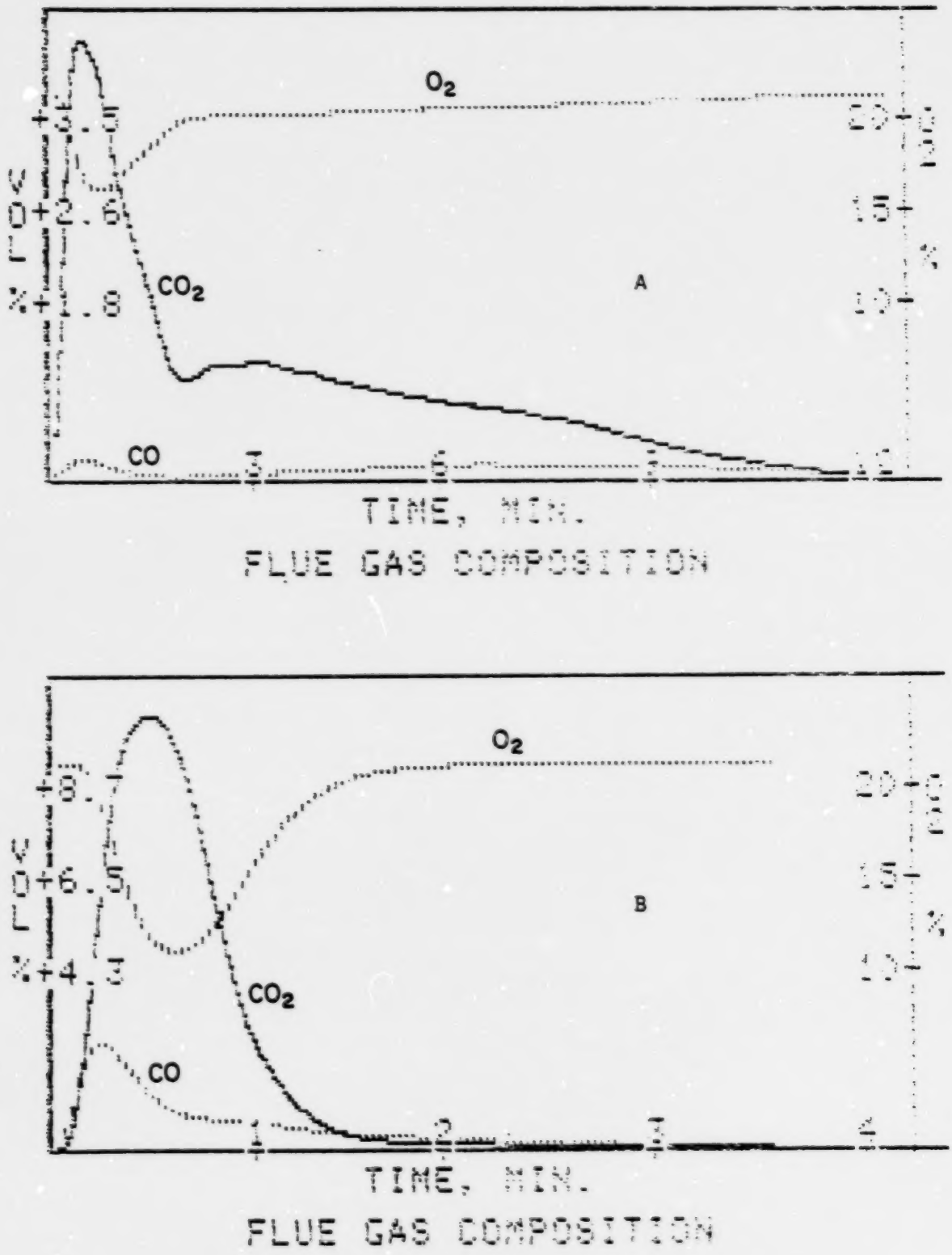

Fig. 4 - Typical flue gas composition

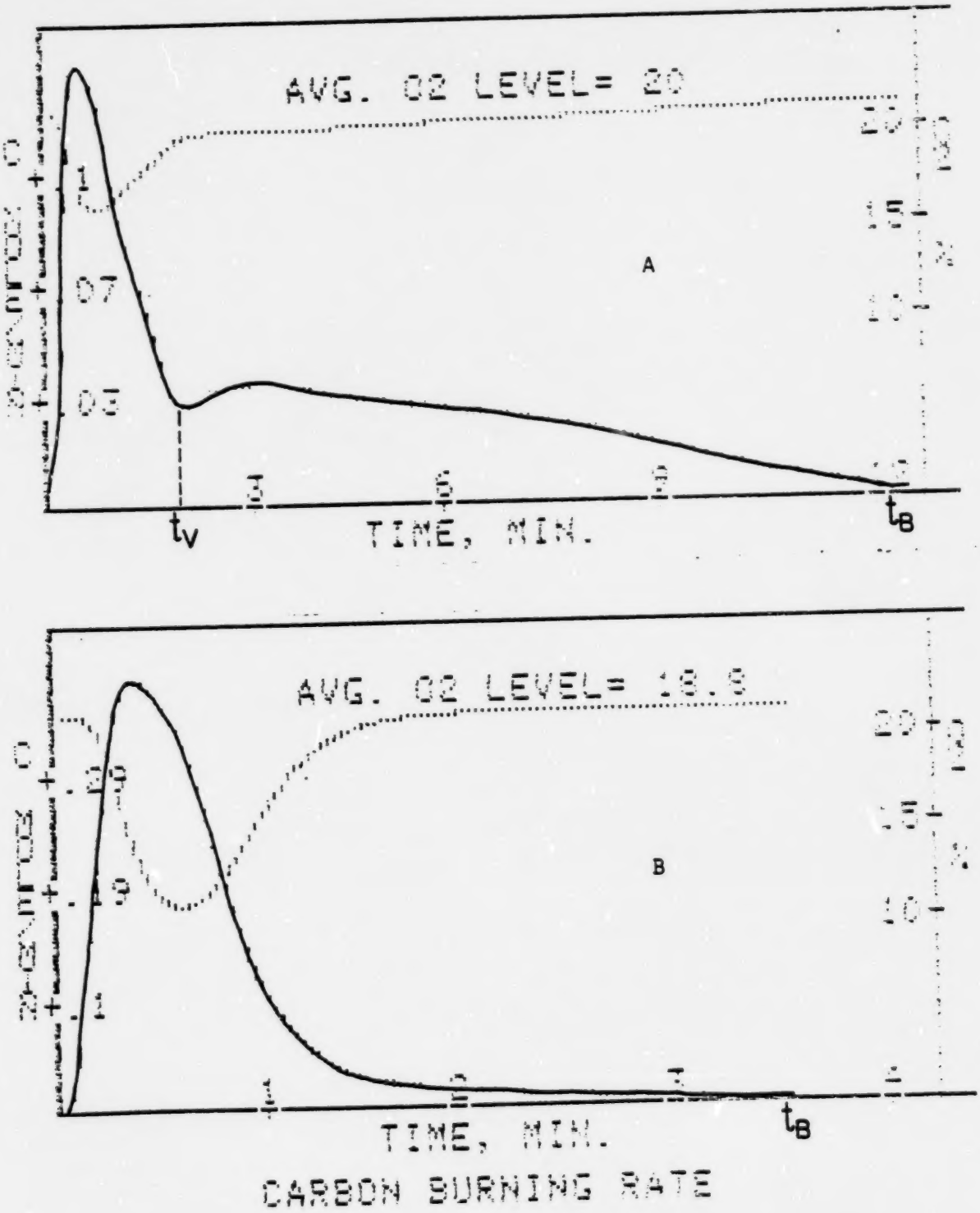

Fig. 5 - Typical carbon burning rate

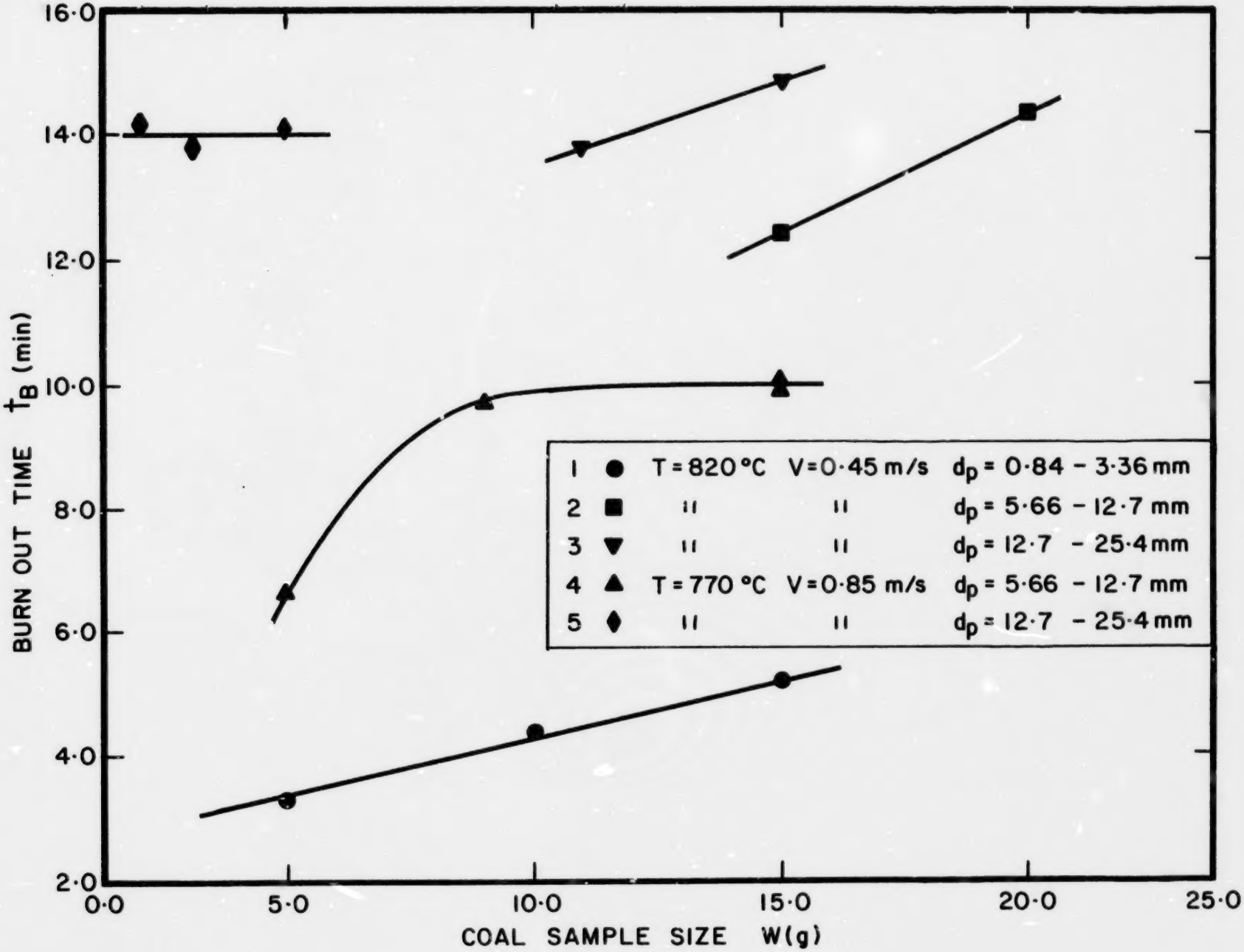

Fig. 6 – Effect of coal sample size

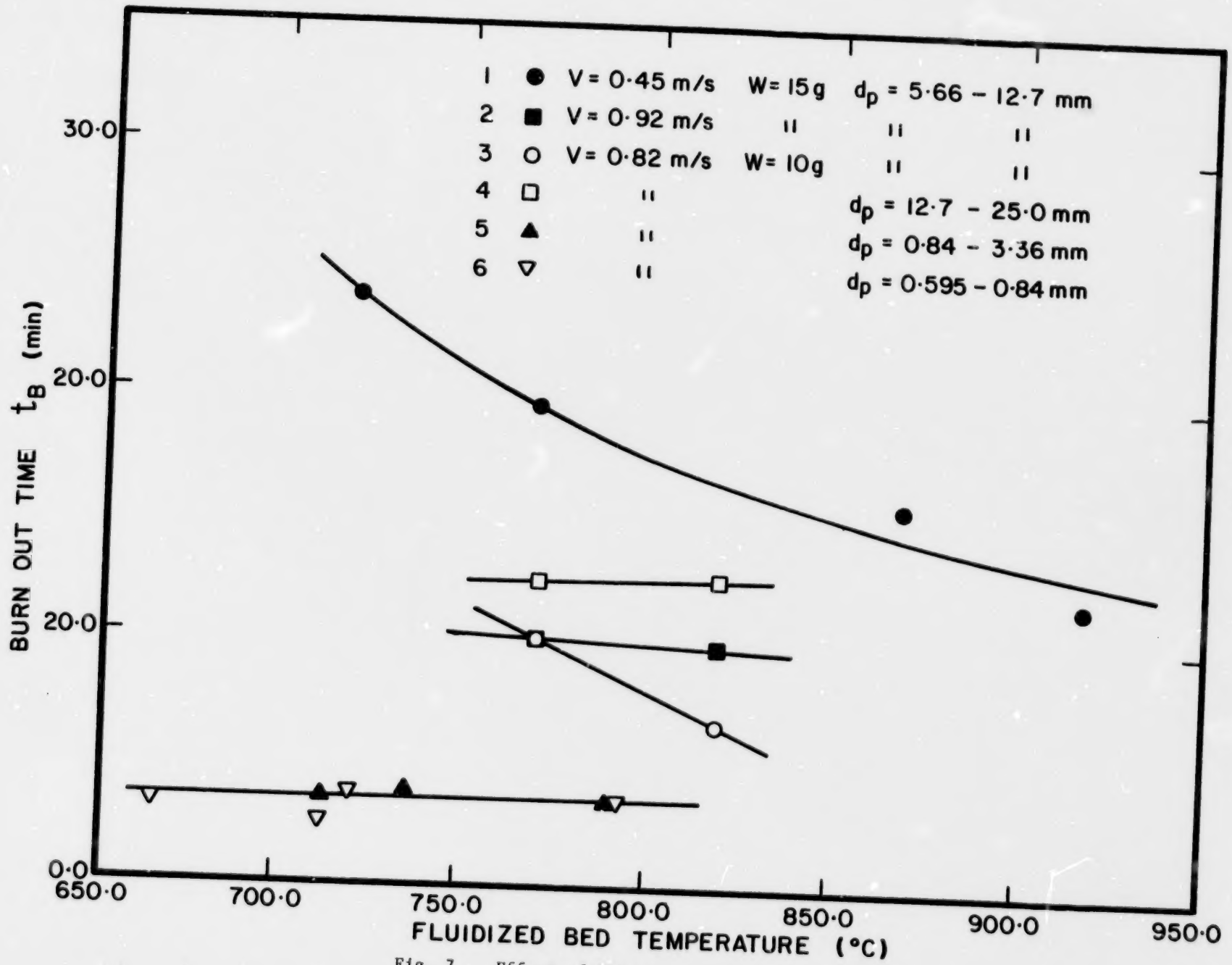

Fig. 7 – Effect of bed temperature

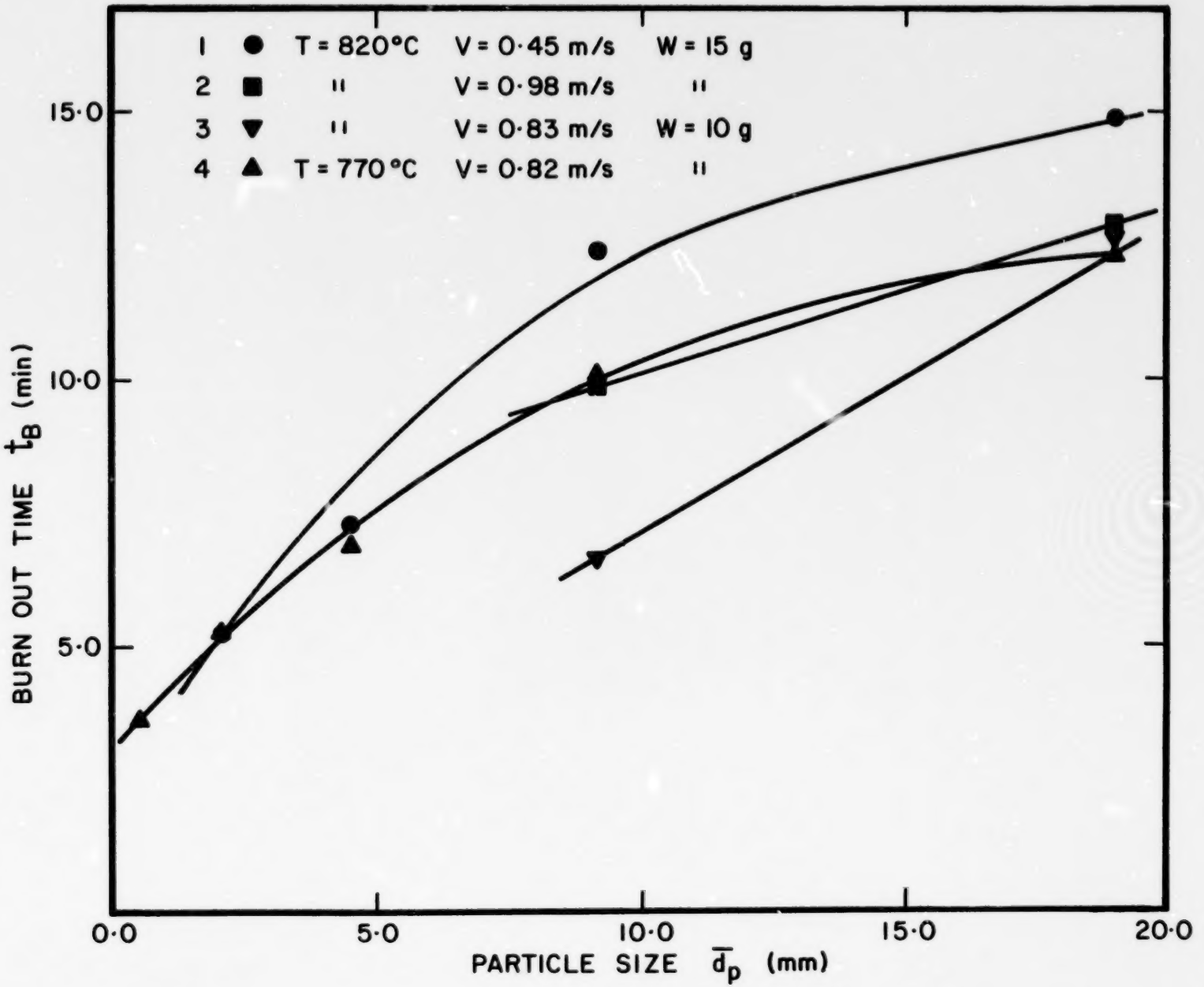
1 ● T = 820°C V = 0·45 m/s W = 15 g
2 ■ " V = 0·98 m/s "
3 ▼ " V = 0·83 m/s W = 10 g
4 ▲ T = 770°C V = 0·82 m/s "
15·0
10·0
5·0
BURN OUT TIME t_B (min)
0·0
5·0
10·0
15·0
20·0
PARTICLE SIZE $\bar{d}_p$ (mm)

Fig. 8 - Effect of particle size